

Thermal and Catalytic Processes in Petroleum Refining

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Table 4.20 Examples of Needle Coke Production

	Decanted Catalytic Cracking Column Bottom	Tar
Feed		
d_4^{20}	1.082	1.085
S, %	0.5	0.56
Conradson carbon	—	8.6
Yields, wt %		
Gases	9.8	18.1
Gasoline	8.4	0.9
Gas oil	41.6	21.1
Coke	40.2	59.9
Operating conditions		
Furnace outlet, °C	510	502
Pressure in drum, bar	—	3.5
Recycle coefficient	2.0	2.08

4.2.3 Coking on a Heat Carrier

The use of a solid heat carrier has the advantage that it makes possible reaching any conversion of the feedstock to coke and lighter products. Indeed, contrary to delayed coking, the conversion is not limited any more by formation of coke deposits inside the tubes of the furnace and on the adiabatic operation of the cracking chambers.

The reaction temperature is reached by the contact of the feedstock with the heat carrier that was preheated to the necessary temperature. The desired conversion is obtained by controlling the residence time in the reactor. The coke is deposited on the heat carrier, which may be made of a material that is inert towards the reaction or coke particles. The products in the vapor state are separated from the carrier. The carrier particles are sent back to be reheated by the partial or total combustion of the coke deposited on them.

Depending on the type of heat carrier used in the reactor, there are processes with a moving bed of heat carrier and units with a fluidized bed carrier. The latter may be implemented as “flexicoking” where the coke formed in the process is gasified.

Coking in the fluidized bed. The first industrial unit for coking in a fluidized bed had a capacity of 600 t/day, and was put in operation in December 1954. Presently, this process is second after delayed coking, and accounted for 11% of world coke production in 1990.

The basic process flow diagram is given in Figure 4.25.

Variations to this scheme occur mainly in the transport system and with reference to the place where the preheated feedstock is fed: directly in the reactor as in Figure 4.25, or in the lower portion of the column situated above the reactor, from which it goes in the reactor together with the recycling material. The heat carrier is formed by coke particles.

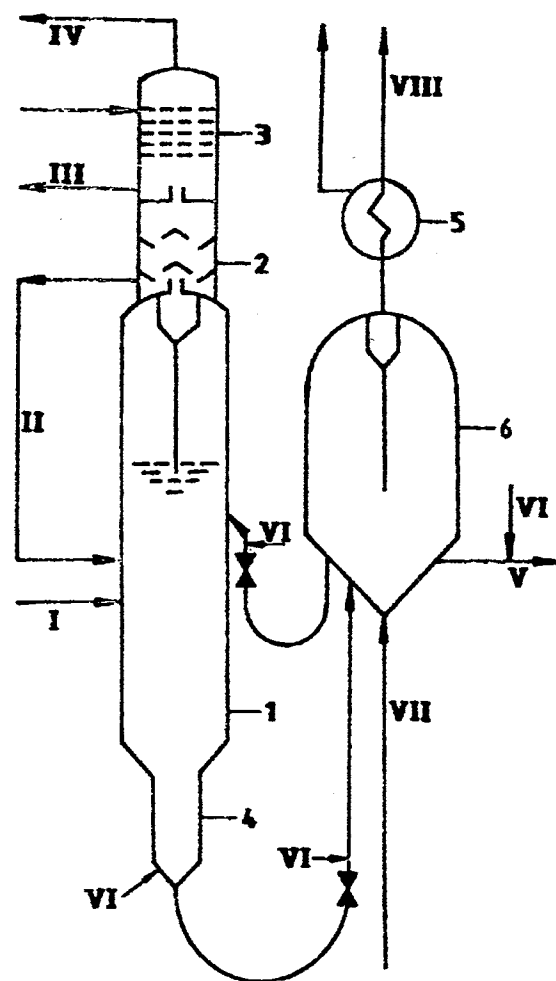


Figure 4.25 Fluidized bed coking. 1—reactor, 2,3—fractionation column, 4—stripper, 5—steam production, 6—heat carrier reheating; I—feed, II—recycle, III—heavy gas oil, IV—to fractionation, V—coke, VI—steam, VII—air for coke burning, VIII—flue gases.

In comparison with other units, using fluidization, coking units have the particular trait that the feedstock, being introduced in the reactor in liquid state, cannot ensure the fluidized state. The fluidization is ensured by the steam introduced in the lower part of the reactor. Its speed in the reaction zone must range between 0.3–0.9 m/sec, depending on the operation conditions, especially reaction temperature [36].

In the upper portion, the reactor has a larger diameter, because through the upper section flows not only the steam but also the vapors of the products formed in the process.

This portion serves as the reaction zone, whereas in the lower portion of the reactor only the steam circulates, performing concomitantly the stripping of volatiles from coke particles.

The cyclones, traditionally used for capturing particles from a gas stream, cannot be used here due to the danger of coking them. The coke particles entrained from the fluidized bed are partly carried in the column that is placed directly above the reactor.

This placement avoids the transfer lines and the danger of their becoming filled with coke. The entrained coke particles are returned back to the reactor together with the recycled material; a minimum flowrate of recycling is necessary for good operation of the unit.

The column is provided in the lower section with baffles and in the upper section with fractionating trays.

In order to obtain coke with a reduced content of volatiles, the residence time of the feed in the reactor is 15–20 sec. However, even with this residence time the coke remains with a residual content of 1–2% of volatiles, which requires an additional calcination. This occurrence is due to the backmixing present in the fluidized bed and leads to some particles leaving the reactor before the coke formation process is completed.

The rate of feed supplied to the reactor, expressed as kg per hour and per kg of heat carrier circulating through the reactor, depends on the coke content of the feedstock, on the temperature, on the recycle rate of the heavy products from the column, and on the velocity of the steam through the reactor.

An improper control of the parameters leads to an increasing amount of sticky products, which are intermediaries in the formation of coke, the agglomeration of the particles of coke, and finally to the blocking of the reactor.

The estimation of the feedrate, as a function of the reactor temperature and the content of coke in the feedstock, may be performed by means of the graph of Figure 4.26 [45].

In this graph, the feedrate is expressed by the ratio of the hourly circulation rates of the feed and of the heat carrier (coke).

A high steam velocity in the reactor may excessively shorten the contact time, and make it become insufficient for achieving the desired conversion. For this reason, steam velocity of the order of 0.9 m/sec, which is needed in order to maintain the reactor diameter at reasonable values, will require quite high reaction temperatures. For operating at 510–525°C, the vapors velocity must be maintained within the limits 0.15–0.30 m/sec.

It seems that best results are obtained at temperatures of 540–550°C and at a velocity of 0.3 m/sec.

At reaction temperatures over 565°C, gas oil becomes excessively aromatized, and its use is difficult without a deep hydrotreatment.

The results obtained in coking depend to a great extent on the recycling of the heavy products formed in the process.

In normal operating conditions the fractions boiling above 545°C are recycled and the coke formed in the process represents 110–130% of the Conradson carbon of the feedstock. This percent decreases sensibly if recycling is zero [36]. The improvement of the quality of the heavy distillate, in view of its use in good conditions in the catalytic cracking process (d_4^{20} between 0.876 and 0.904, coke under 0.3%), imposes a high recycle ratio. This fact however diminishes the capacity of the unit. The recycled substance, being more stable as the feedstock, needs a somewhat longer residence time in the reactor or a temperature situated towards the highest admissible limit.

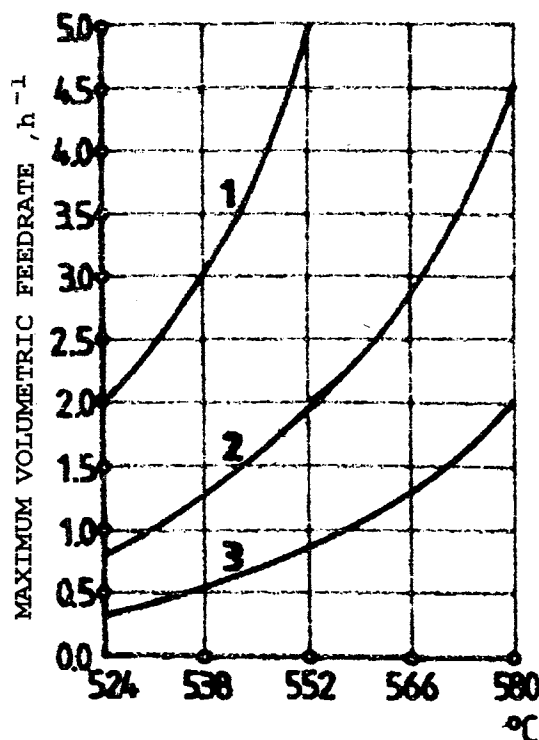


Figure 4.26 The maximum feeding velocity function of the reactor temperature, for three feed qualities. 1—Feed with Conradson carbon = 12%, 2—Same = 22%, 3—Same = 29%.

The coke particles are reheated by burning a portion of the coke in a fluidized bed furnace. Laboratory tests using a fluidized bed with a depth of 0.15–0.25 m, with 90% of the particles having diameters comprised between 0.25–0.67 mm, were reported. Operating the fluidized bed at the temperature of 600°C, with superficial gas velocities of 0.5–0.8 m/s, burning rates of 270–300 kg/hour·m³ were measured [46].

In commercial units, the height of the fluidized bed is of 3–4 m, and the burning rate of coke is somewhat lower.

The mechanical qualities of the coke do not change if the residence time in the furnace is between 5–10 minutes. At excessively long residence times the particles become brittle and their volume density decrease from 1.030 to 0.805 g/cm³.

The coke burnt for preheating represents generally about 5% of the feedstock. The rest, preferably the larger particles, is removed from the system.

The yield and the quality of the products obtained by the fluidized bed coking of various feedstocks, with recycling of the fractions with boiling points above 545°C, are given in Table 4.21.

Typical characteristics of the coke are: 0.6% volatiles at 593°C and 5.0% at 950°C; hydrogen content in the coke is 1.6%.

Table 4.21 Fluidized Bed Coking of Various Feedstocks

	Feedstocks				
	I	II	III	IV	V
Density	1.0427	0.9509	0.9484	1.0772	1.1010
Conradson carbon	24.5	11.0	5.0	33.0	41.0
Sulfur, %	4.3	0.7	0.5	2.3	2.1
Nitrogen, %	0.28	0.32	0.2	1.9	2.1
C/H ratio	8.4	6.2	7.0	9.2	9.9
Viscosity at 135°C, cSt	88	55	—	—	—
Distillation at 538°C, % vol.	10	0.0	40	0.0	5.0
Gases C ₃ , wt %	9.5	7.0	6.0	10.0	11.5
Overall gasoline, wt %	27.5	11.5	8.0	36.0	48.5
C ₄ , % vol.	3.5	2.0	2.0	3.0	3.0
Debutanized gasoline, fbp 221, % vol.	19.5	21	17	17.5	14.5
Gas oil 221–546°C, % vol.	52	68.5	74	44.5	32.5
Gasoline					
Density	0.7547	0.7425	0.7547	0.7628	0.7587
Octane number F1	77	66	73	82	—
Sulfur, wt %	0.9	0.2	0.1	1.7	0.8
Gas oil					
Density	0.9529	0.8816	0.9100	0.9729	0.9659
Conradson carbon	2.8	1.2	0.8	3.1	1.6
Sulfur, %	3.7	0.4	0.4	2.2	1.7
K _{UOP}	11.17	11.98	11.81	10.97	11.01
Temperature 50% dist.	416	416	453	416	423

The yields of coke and gases for coking in a fluidized bed may be estimated by means of the relations [36]:

$$C = 1.15c$$

$$C + G = 5.0 + 1.30c \quad (4.1)$$

In these relations, C and G represent the percent of coke and the percent of gases by weight, and c , the content of Conradson carbon in the feedstock.

The relations (4.1) are approximate since they do not take into account the recycle ratio.

More exact results are obtained by using Figures 4.27 and 4.28 [48]. Figure 4.27 allows the determination of the yields of coke and gas oil and of their main characteristics.

When the feed is straight run residue, the yield of debutanized gasoline represents 17% (density of gasoline 0.755) and when feeding vacuum residue, it represents 21% (density of gasoline 0.765) of the product.

The yield of the C₁–C₃ fraction is calculated by difference.

Figure 4.28 gives the octane number of the gasoline, the distribution of sulfur among the fractions, and the yield of C₄ fraction.

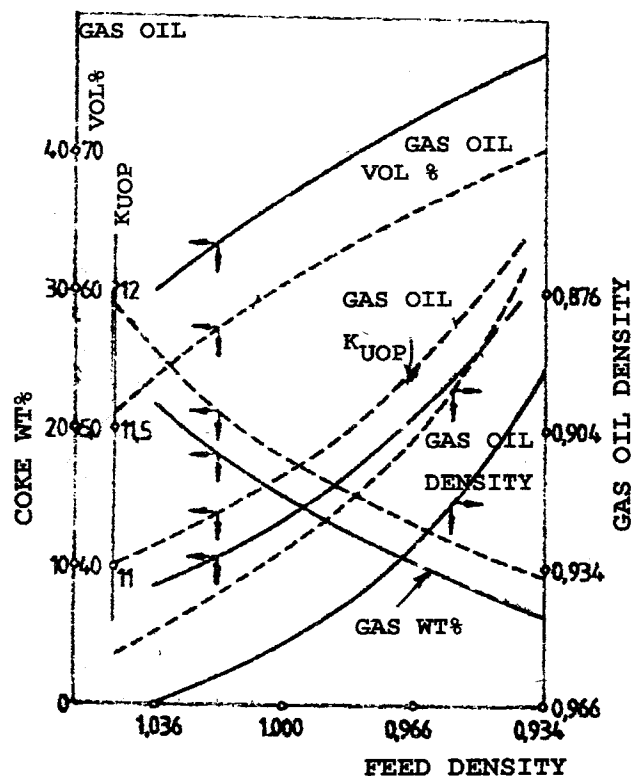


Figure 4.27 Gas oil and coke yields and density in fluid coking. (From Ref. 48). Continuous line: straight run residue, dotted line: vacuum, thermal cracking or deasphalting residue.

The investments for fluid coking are approximately the same as for the delayed coking, but operating costs are higher [40].

Flexicoking. Flexicoking is a more recent variation (1976) of coking in a fluidized bed. It is the combination between the classic fluidized bed coking and the gasification of the produced coke. In 1990 the process accounted for 1% of world coking capacity.

The typical simplified process flow diagram is given in Figure 4.29. This process flow diagram depicts both the classic system, with a single gasifier (without the part delimited by the dotted line) and the Dual system, provided with two gasifiers. In the second gasifier, the gasification is performed with steam and synthesis gas is obtained [49].

In both versions (classic and dual), the reheating of the coke in the dedicated equipment is not done by injection of air and partial burning of coke, but instead by contact with gases emitted from the gasifier. In the gasifier, partial burning of coke is carried out with a mixture of air and steam, and fuel gas is obtained. When a second gasifier is added, the coke from the first gasifier is sent to the second, where the

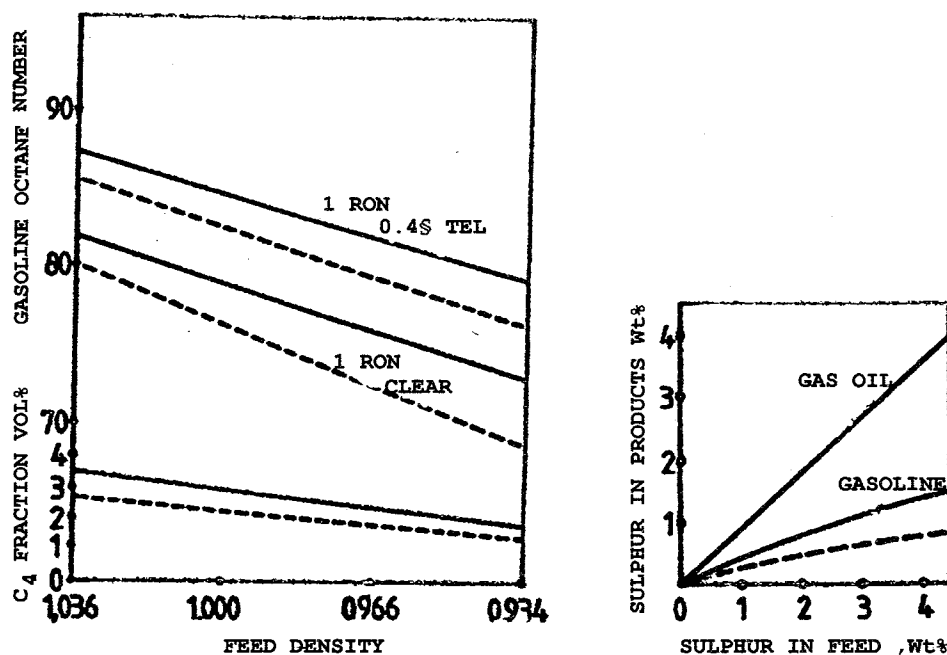


Figure 4.28 Gasoline octane number, C₄ yield, sulfur distribution in fluid coking. Continuous line: straight run residue, dotted line: vacuum, thermal cracking or desasphalting residue. (From Ref. 48.)

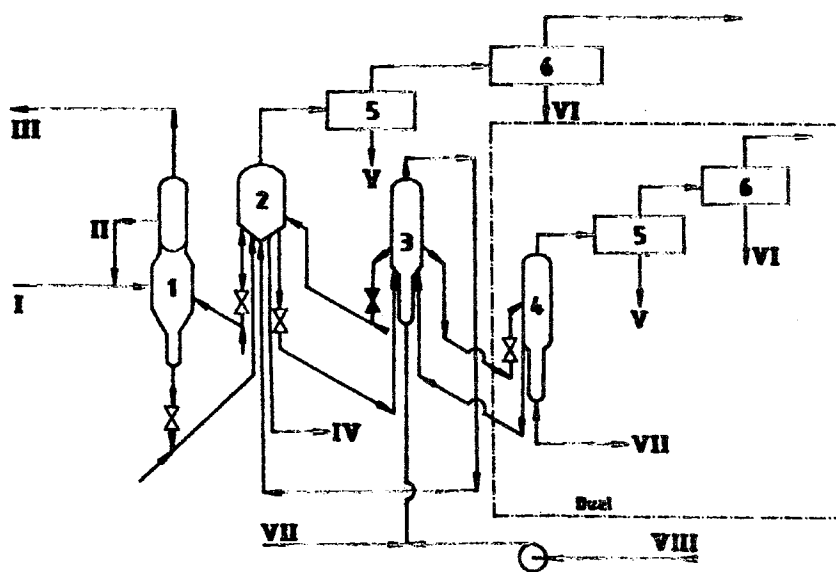


Figure 4.29 Flexicoking process flow diagram. 1—reactor, 2—carrier reheating, 3—gasifier, 4—“dual” gasifier, 5—coke dust removal, 6—desulfurization; I—feed, II—recycle, III—vapors to fractionation, IV—coke, V—coke dust, VI—sulfur, VII—steam, VIII—air.

synthesis gas is obtained. This may be used for the synthesis of methanol or of hydrogen, for the production of ammonia.

Typical yields and economics for the fluid coking and flexicoking processes are given in Table 4.22 [109].

A comparison between the yields obtained by delayed-, fluid- and flexicoking of the same feedstock, is given in J.H. Gary and G.E. Handwerk's monograph [101].

Moving bed systems. The initial attempts to develop systems with moving beds of coke particles, or inert material encountered great difficulties due especially to coke depositing on the walls of the reactor and eventually blocking circulation. In order to eliminate this phenomenon, a very high carrier/feed circulation ratio and other measures were needed, which made the system to be uneconomical [3].

Recently, the solution to these difficulties was obtained by shaping the reactor as a horizontal cylinder. The coke particles enter at a temperature of 600–700°C and are moved by means of a screw conveyor, whereby they are brought into contact with the feedstock. The screw conveyor hinders the formation of coke deposits on the wall. The coking takes place at temperatures of 500–600°C.

Table 4.22 Typical Fluid Coking and Flexicoking Data

Feedstock		
°API		3.2
Conradson carbon		28.5
Sulfur, wt %		5.6
Vacuum distillate fbp 1050°F from heavy Saudi Arabian crude		
Yields		
Gases C ₁ –C ₄ , wt %		12.9
C ₅ –430°F, % vol		14.4
430–650°F, % vol		10.2
650–975°F, % vol		27.3
For fluid coking, net coke yield, tons/bbl		0.05
For flexicoking low Btu gas, 10 ³ BTU/bbl		1,320
	Flexicoking	Fluid Coking
Economics		
Investment 96, \$/bpsd	2,400–3,100	1,600–2,100
Utilities per bbl feed		
Electricity, kWh	30	30
Steam, 125 psig, lb	150	25
Water, boiler feed, gal	35	35
Water, cooling, gal	700	45
Steam 600 psig (produced), lb	(200)	(160)
Air blower		
Compressor HP-hr	0.6	0.2
Steam 125 psig (produced), lb	(660)	(230)
Steam 600 psig, lb	660	230

Installation: Until 1994, 5 flexicoking units built, over 160 Kbps capacity; Licensor: Exxon Research & Engineering Co.

Source: Ref. 109.

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The flow diagram of the plant, known under the name of *Coking-LR* is presented in Figure 4.30.

The burning of coke and reheating of the coke granules take place in a vertical transport pipe.

The unit converts not only heavy residues obtained in crude oil processing but also natural solid bitumen or other similar feedstocks.

Although the construction of several pilot plants and also commercial scale units was reported [44], no information concerning capacities and plant performance is available.

Another coking process with solid heat carrier which was proposed as early as 1981 and is known as *Dynacracking*. The process uses alumina as heat carrier.

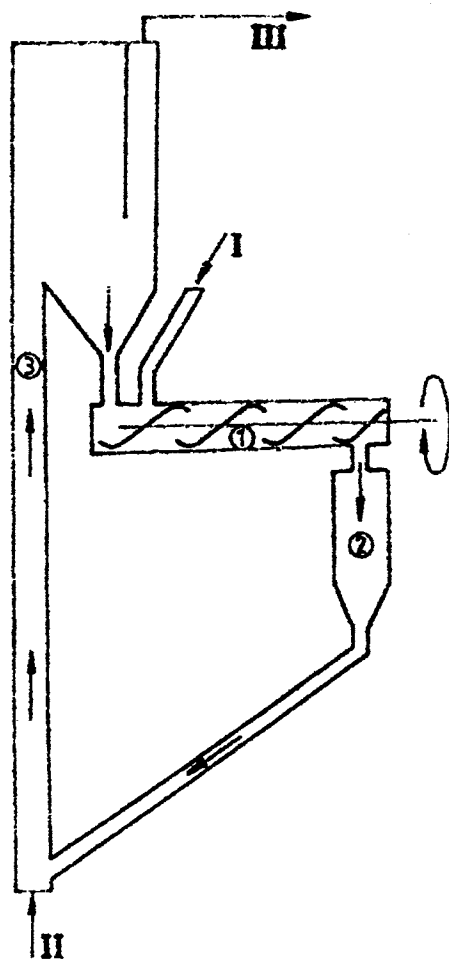
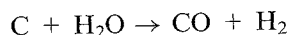


Figure 4.30 Moving bed LR coking unit. 1—reactor, 2—buffer vessel, 3—riser and re-heater; I—feed, II—air, III—flue gases.

The burning of coke uses a mixture of oxygen and steam, their proportions being controlled so that the following reaction is favored:



and the consumption of oxygen is minimized [51].

The produced gases are rich in hydrogen. As they pass through the reaction zone, coking takes place in the presence of hydrogen at a hydrogen partial pressure of 5–15 bar. In this way, the amount of coke decreases and the quality of the products is favorably influenced.

Other proposed processes that have not yet become popular in industry are described summarily [6,12].

4.2.4 Coke Calcination

Coke calcination is carried out mostly in tilted rotary cylindrical furnaces similar to those used in the cement industry. In the past, such operations were competitive only at very high capacities. For this reason the units were built as separate entities that processed the coke produced by several refineries. The subsequent development of plants of lower capacities (40,000–300,000 t/year) [53] led in the last two decades to the trend to perform coke calcination in the refineries themselves.

Calcination is used for treating the spongy coke obtained in the delayed coking, the needle coke, and the coke obtained in the fluid coking.

The calcination of needle coke produces graphite, which is used for the construction of special equipment (heat exchangers, pumps, valves, chemical reactors, crucibles, pipes a.o.) in the nuclear industry, construction of electrical motors, in the fabrication of electrodes for the caustic and chlorine industry, and for electrical metallurgical furnaces.

The coke obtained by calcining that was produced in delayed coking and in fluid coking is used for the fabrication of anodes for the aluminum industry and in some electrometallurgical and electrochemical industries. In the production of aluminum the consumption of coke is of 0.4–0.5 kg coke/kg aluminum. Since the world production of aluminum is 17–20 mill. t/year (1985), this industry alone consumes yearly 8–10 million tons of calcined coke.

Typical characteristics of raw coke and specifications for the calcined coke are given in Table 4.23.

Tilted rotary furnaces are used in about 95% of plants for calcination of petroleum coke. See Figure 4.31.

The heat required for calcination is supplied by the burning of volatile substances contained in the coke and of gaseous, liquid, or solid additional fuel as well as by the calcination of small amounts of coke. The complete combustion of the volatile substances is ensured by the secondary and tertiary air introduced into the furnace.

The calcined coke is discharged in a rotary cooler where the temperature is reduced to 90–120°C. Various cooling systems are in operation: water pulverization, air preheating, indirect cooling etc.

After being discharged from the cooler, and before it is sent to storage, an oil is usually pulverized over the coke, in order to reduce the dust produced during handling.

5.3.5 Design Elements for Fluidized Bed Coking Units

Backmixing that occurs in the dense phase fluidized bed is very intensive and the reactor and coke burning bed approach a perfectly mixed reactor. Kinetic calculations must take this into account.

Backmixing also has as a result the homogenization of the temperature inside the dense phase fluidized bed. The difference between different points of the bed does not exceed 1–2°C.

The reactor and the heater in coking and flexicoking plants used dense phase fluidized beds. In these conditions, the reaction temperature inside the bed may be easily correlated with the temperature of the two streams, solid and fluid, entering the reactor, using in this purpose the thermal balance of the reactor*.

$$G_{mp}[\bar{c}_p(t_r - t_{im}) - q_r x] + G_s c_s(t_r - t_{is}) + Q_p = 0 \quad (5.65)$$

When the solid is a catalyst, the reactions will cease altogether at the separation of the solid from the fluid. In the case of thermal (noncatalytic) processes, they will go on as long as the effluent is not quenched when leaving the reactor. This fact must be taken into account in the design of the process.

Two systems are used in thermal cracking for transporting the solids between the reactor and the heater:

The ascendant transport at large dilutions of the solid similar to the pneumatic transport

The transport in fluidized dense phase, the fluidized state being maintained by injections of inert fluid from place to place, along the transport pipe

The first transport system—the pneumatic transport—is a classic, well-known system that does not need any further explanations.

The second system is based on the fact that a homogeneous fluidized bed in a dense phase behaves in some respects as a liquid. The bed flows under the action of differences in the hydrostatic level or of a pressure gradient provided that the fluidization is maintained. Generally, this system is used in combination with the pneumatic transport.

The main advantage of the system that transports solids in a dense fluidized phase is the practical elimination of the erosion of the transport pipes and of the solid particles. These advantages appear especially obvious in the points of the unit where the transport pipes change direction and where strong erosions are observed as a result of operation in the pneumatic transport mode. For this reason, in the combined systems the pneumatic transport is used as much as possible through ascending straight pipes.

The design of the system depends on the particular process and may meet additional difficulties, such as in the case of fluid coking, when a liquid phase is injected into the bed. In this case the theoretical calculation is very difficult and uncertain.

In fluid coking, the amount of coke deposited on the coke particles in the reactor exceeds the amount that must be burnt in order to reheat the coke particles.

Accordingly, the thermal balance of the reheating furnace has the form:

* The thermal balance of the furnace used for reheating is carried out in a similar manner.

$$\begin{aligned}
 G_s c_s (t_{ec} - t_{ic}) + G_a \bar{c}_{pa} (t_{ec} - t_{ia}) + Q_p \\
 = G_{mp} x_c q_a + G_g \bar{c}_{pg} t_{ec} + G_{cb} q_{cb}
 \end{aligned}
 \quad (5.66)$$

where:

- x_c = coke burnt for reheating, reported to the feed
- q = the thermal effects
- Q_p = the losses of heat through the walls
- subscript s = heat carrier
- a = air,
- mp = feed
- g = burnt gases
- cb = the supplementary fuel
- ic = the inlet in the furnace
- ec = outlet from the furnace
- ia = inlet of air

The fact that the heat losses of the transport system are not important and may be estimated to cause a temperature decrease of 10–20°C allows the correlation of the inlet and outlet temperatures of the heat carrier in the reactor and in the reheater. Thus, for the correlation of Eqs. (5.65) and (5.66), the following values are accepted:

For the transport of the reheated carrier:

$$t_{ec} - t_{is} = 15 - 20^\circ\text{C} \quad (5.67)$$

For the transport of the carrier that left the dense phase reactor:

$$t_{ir} - t_{ic} = 10 - 15^\circ\text{C} \quad (5.68)$$

The decrease of temperature is higher for the transport of the reheated carrier. This is justified by its much higher temperature. A final, exact calculation will specify the Δt values.

A last problem concerning the design of the reactor-reheater system is the correlation of the pressures in these two vessels and of their relative height in order to ensure a correct circulation of the heat carrier (coke particles).

The calculation depends on whether the transport is performed in dense phase through semicircular pipes and in diluted phase only in a part of straight ascendant pipes (coking plant, Figure 4.23) or through straight ascendant and descendent pipes (Figure 4.27). In the second case the solids are in diluted phase in the ascendant pipes and in dense phase in the descendent ones.

In both cases, the pressure in the reactor is determined by the hydrostatic and hydrodynamic pressure drops in the fractionation system and by the aspiration pressure of the gas compressor.

The pressure in the reheating furnace is determined by the pressure drop through the heat recovery system and cleaning of the flue gases.

The height and pressure parameters that play a part in the design of the transport of solids through semicircular pipes is given in Figure 5.8.

The gas injections in the semicircular portions of the pipes have the role of maintaining the state of dense phase fluidization. The valves located on the two pipes serve exclusively for the isolation, in case of need, of the two vessels. In normal operation they are in the fully open position.

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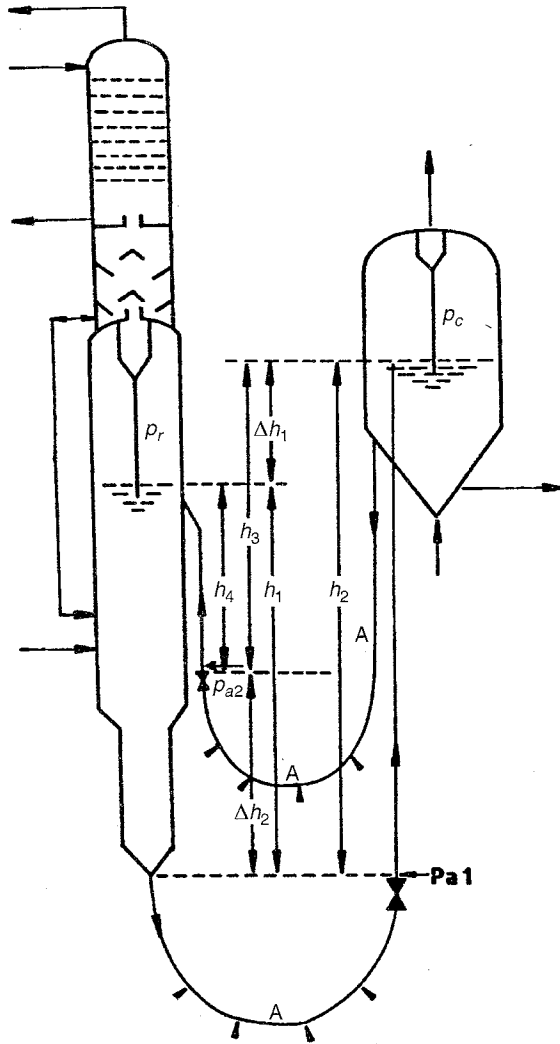


Figure 5.8 Circulation of coke particles in a transport system with curved pipes.

For simplification: the hydrodynamic pressure drop in the pipes for dense phase transport can be neglected, the bulk densities in the pipes for dense phase transport can be considered identical. The same is valid for the pipes for dilute phase transport.

The conditions for accurate circulation may be expressed by the following equations:

$$\left. \begin{aligned} p_{a1} &\geq h_2 \gamma_{\text{dil}} + p_c + \Delta p_2 \\ p_{a1} &< h_1 \gamma_{\text{dens}} + p_r \\ p_{a2} &\geq h_4 \gamma_{\text{dil}} + p_r + \Delta p_4 \\ p_{a2} &< h_3 \gamma_{\text{dens}} + p_c \end{aligned} \right\} \quad (5.69)$$

where:

p_{a_1} and p_{a_2} = pressures at the feed points of the transport agent, indicated in Figure 5.8,

Δp = hydrodynamic pressure drops

γ_{dens} and γ_{dil} = the densities of the dense and of the diluted phases

The rest of the notations are shown in the same figure.

Equating these equations pairwise, one obtains:

$$\left. \begin{aligned} h_2 \gamma_{\text{dil}} + p_c + \Delta p_2 &= h_1 \gamma_{\text{dens}} + p_r \\ h_4 \gamma_{\text{dil}} + p_r + \Delta p_4 &= h_3 \gamma_{\text{dens}} + p_c \end{aligned} \right\} \quad (5.70)$$

By subtracting the second equation from the first, it results:

$$(h_2 - h_4) \gamma_{\text{dil}} + p_c - p_r + \Delta p_2 - \Delta p_4 = (h_1 - h_3) \gamma_{\text{dens}} + p_r - p_c \quad (5.71)$$

According to the figure,

$$h_2 - h_4 = \Delta h_1 + \Delta h_2$$

and

$$h_1 - h_3 = \Delta h_2 - \Delta h_1$$

By substituting and regrouping the terms of the last expression, one obtains:

$$2(p_r - p_c) = \Delta h_1 (\gamma_{\text{dens}} + \gamma_{\text{dil}}) - \Delta h_2 (\gamma_{\text{dens}} - \gamma_{\text{dil}}) + \Delta p_2 - \Delta p_4 \quad (5.72)$$

As the pressures in the two vessels and the bulk density in the dense phase are imposed by the process conditions, the correct circulation is ensured by the corresponding selection of the relative height of the two vessels— Δh_1 , and of the position of the transport pipes that determines the distance— Δh_2 . To a lesser extent it is possible to change the density, γ_{dil} , of the diluted phase. Its selection is determined by considerations related to the erosion of the ascending transport pipes.

Remark: in this transport system, accidental pressure increase in one of the vessels cannot lead to the inversion of circulation; even a beginning of reverse circulation leads to the blocking of the semicircular pipes by changing the transported material from the fluidized state to a settled bed.

In the transport system with straight pipes (Figure 5.9) solids transport is not blocked if a sudden increase of pressure occurs in one of the vessels. To absorb such shocks and to prevent the inversion of circulation, the two valves situated at the lower ends of the descendent transport pipes in dense phase are controlled to ensure an approximately 0.2–0.3 bar pressure drop in normal operation. An accidental increase of the pressure in one of the vessels diminishes this pressure drop, thus preventing the inversion of the circulation.

The equations that express the conditions of proper circulation will have in this case the form:

$$\left. \begin{aligned} p_{a_1} &\geq h_2 \gamma_{\text{dil}} + p_c + \Delta p_2 \\ p_{a_1} &< h_1 \gamma_{\text{dens}} + p_r - \Delta p_{v_1} \\ p_{a_2} &\geq h_4 \gamma_{\text{dil}} + p_r + \Delta p_4 \\ p_{a_2} &< h_3 \gamma_{\text{dens}} + p_c - \Delta p_{v_2} \end{aligned} \right\} \quad (5.73)$$

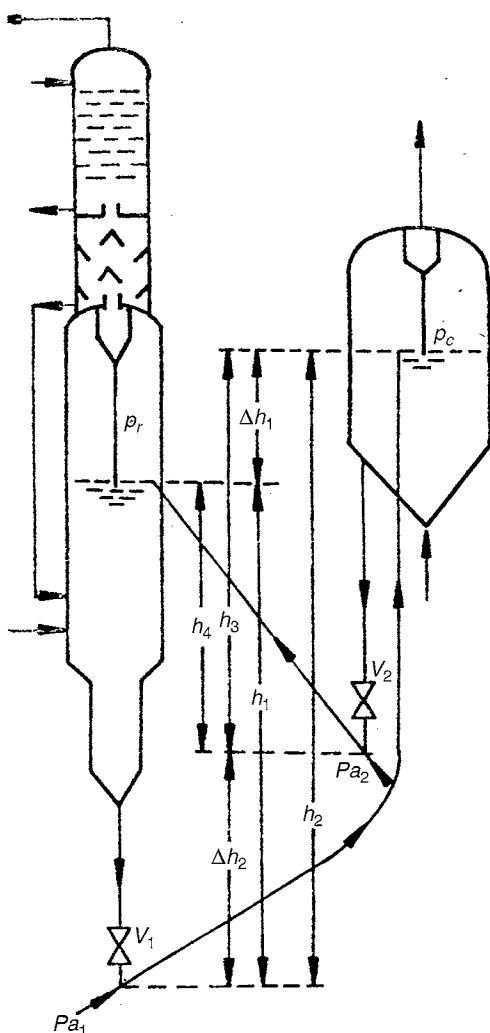


Figure 5.9 Circulation of coke particles in a transport system with straight pipes

By equating the equations pairwise, it follows:

$$\left. \begin{aligned} h_2 \gamma_{\text{dil}} + p_c + \Delta p_2 &= h_1 \gamma_{\text{dens}} + p_r - \Delta p_{v_1} \\ h_4 \gamma_{\text{dil}} + p_r + \Delta p_4 &= h_3 \gamma_{\text{dens}} + p_c - \Delta p_{v_2} \end{aligned} \right\} \quad (5.74)$$

Finally, operating as in the previous case, it results:

$$\begin{aligned} 2(p_r - p_c) &= \Delta h_1(\gamma_{\text{dens}} + \gamma_{\text{dil}}) - \Delta h_2(\gamma_{\text{dens}} - \gamma_{\text{dil}}) + \Delta p_2 - \Delta p_4 + \\ &+ \Delta p_{v_1} - \Delta p_{v_2} \end{aligned} \quad (5.75)$$

If, as in the case of normal operation the pressure drops through the two valves are equal, Eq. (5.75) becomes identical to (5.72).

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